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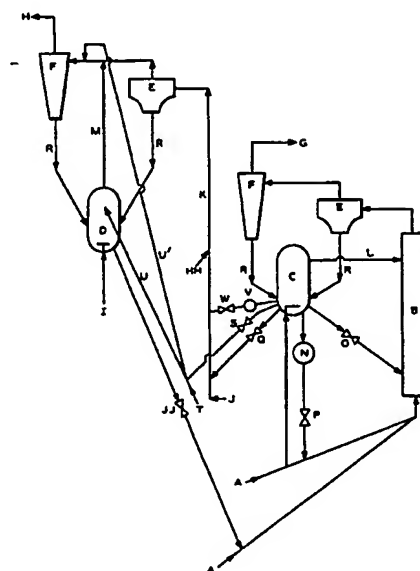
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(54) Hydrocarbon treatment process.

(57) A process wherein lower boiling products (H) are obtained from a hydrocarbon feed (J and/or HH) by contacting the feed with fluidized solid contact material at elevated temperature in a riser (K), spent contact material is separated and stripped of volatile hydrocarbons in a stripping zone (D), stripped contact material is regenerated with oxygen-containing gas (A) in a regeneration zone (B), and hot freshly regenerated fluidized solid contact material is returned to the riser (K), the process further including two or more of the following:

- (i) suspending hot regenerated contact material in a carbonizable lift gas at the lower portion of the riser and injecting hydrocarbon feed higher in the riser;
- (ii) charging hydrocarbon feed into the base of the riser, steam cooling a portion of the hot freshly regenerated contact material, and contacting the hydrocarbon feed with hot regenerated contact material upstream of where said hydrocarbon feed is contacted with the steam cooled contact material;
- (iii) cooling a portion of hot freshly regenerated contact material before returning it to the riser; and
- (iv) adding a portion of freshly heated regenerated contact material to said stripping zone.



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HYDROCARBON TREATMENT PROCESS

This invention relates to a process and system for obtaining light fractions from heavy hydrocarbon oils, including processes for selective vaporisation and processes for converting heavy hydrocarbons containing high concentrations of coke precursors and heavy metals into gasoline and other liquid hydrocarbons. In a preferred embodiment this invention is directed towards the fluid catalytic cracking of hydrocarbons to obtain products boiling in the motor fuel range.

The fluid catalytic cracking process, with the advent of the highly active zeolitic type catalysts, has evolved into the generally standardized practice of effecting essentially cocurrent ascending flow of hydrocarbon vapors and the finely divided catalyst in an elongated tubular reaction zone referred to in the industry as a riser. Notwithstanding the brevity of the reaction cycle, which is usually in the order of about 10 seconds or less, there is a laying down of coke on the catalyst thereby adversely affecting its activity as well as

1 undesirably altering product yield distribution upon reuse.
Accordingly, it is the universally observed procedure to
separate the catalyst from the riser effluent and recycle it
to the process via a regenerator which serves the dual
5 purpose of combusting the coke contaminants and heating the
catalyst for reuse in the reaction cycle.

In accordance with the prior art, separation of
the catalyst from the cracked hydrocarbons is carried out
within a so-called disengaging chamber which contains a
10 stripping zone. The disengaging chamber is a contained
vessel either forming a relatively voluminous shroud about
the downstream extremity portion of the riser or externally
positioned and axially aligned therewith. In the latter
type arrangement, the fluid stream of catalyst and
15 converted hydrocarbons is discharged into the disengaging
chamber directly from the riser via a sidewise opening or
port. In said shroud-type arrangement, however, it is
generally preferred to pass the riser effluent from said
sidewise opening or port firstly through a singlestage
20 cyclone vented to the disengaging chamber. Due to the
substantial reduction of the superficial space velocity
experienced in the disengaging chamber in either of said
modes of operation, a considerable portion of catalyst
entrained in the riser effluent settles out and collects at
25 the bottom of the chamber. The gas stream along with the
stripping vapor thereupon is vented to the fractionator
from disengaging chamber via a cyclonic separator
positioned therein serving to collect predominantly all of
the entrained catalyst.

30 In light of the fact that the modern zeolitic
cracking catalysts range in particle size from about 5 to
100 microns with the major portions thereof being in the
order of from about 40 to 80 microns, separation thereof as
practiced in accordance with the aforesaid prior art is
35 nonetheless remarkably efficient. However, the relatively

small amount of catalyst entrained in the cracked stream poses a problem because of the tremendous cumulative throughput thereof. In some cases, the foremost problem is that the entrained catalyst leaving with the cracked stream must be recycled thereto in the form of a slurry oil recovered from the fractionator thereby reducing the amount of feedstock that can be accommodated in the cracking unit. Thus, optimal processing efficiency of the cracker is sacrificed.

10 Additionally, there has always been a need in the prior art to affect rapid disengagement of the catalyst from reaction products in order to minimize undesirable reactions which can take place thereby detracting from the overall efficiency of the process when considered as a
15 whole.

 As can well be appreciated by those skilled in the art, a necessary and integral part of a fluid catalytic cracking reactor involves the regenerator wherein the spent catalyst has its activity restored. Regeneration of spent
20 catalyst is generally effected after separation of the spent catalyst from the reaction products. The spent catalyst is removed from the reaction zone and contacted in a stripping zone with a stripping medium, usually steam, to remove vaporized and entrained and/or occluded hydrocarbons
25 from the catalyst. From the stripping zone, a stripped catalyst is passed into a regeneration zone wherein the stripped spent catalyst is regenerated by burning coke deposits therefrom with an oxygen-containing gas, usually air. The resulting hot regenerated catalyst from the
30 regeneration zone is then recycled to the reaction zone and contacted with additional hydrocarbon feed. The efficiency of stripping affects the heat released in the regenerator. In practice, commercial strippers are not completely efficient and residual hydrocarbon is present in the

1 catalyst discharged therefrom. Regenerator temperatures
increase as the efficiency of stripping decreases.

The ratio of recycled regenerated catalyst to
hydrocarbon feed, referred to generally as the C/O
5 (cat-to-oil) ratio, affects selectivity in a typical FCC
unit. The effect is most pronounced in a heavy oil FCC
unit. The higher the C/O, the better the selectivity and
the lower the contact time can be. Lower contact time
results in lower hydrogen transfer when using zeolitic
10 cracking catalyst. Lower hydrogen transfer also results in
higher gasoline octane, increased olefins for alkylation
feedstock, and higher hydrogen content LCO for distillate
production for a given conversion. In commercial FCC
units, catalyst circulation rate (CCR) and catalyst to oil
15 ratio (C/O) are not independent variables that can be
changed at will. Heat balance considerations in commercial
unit establish the operating C/O. Generally, the only
variables that are independently controlled is cracking
temperature (in particular, the temperature at the outlet
20 of the riser cracker) and feed preheat temperature.

Selective vaporization is carried out in
equipment similar to that used in FCC operations. The
fluid solid contact material, however, is substantially
inert as a cracking catalyst. Selective vaporization
25 occurs in a riser, called a contactor, and combustion of
carbonaceous deposit takes place in a burner. See, for
example, U. S. 4,263,128 (Bartholic) which is herein
incorporated by reference. The technology is known in the
industry as the ART process.

30 The term Δ coke is the weight percent of
coke on spent catalyst minus the weight percent coke on
regenerated catalyst. In other words, Δ coke is the weight
of coke on spent catalyst minus the weight coke on
regenerated catalyst divided by catalyst circulation rate
35 (CCR). Delta coke is related to C/O by the equation:

- 5 -

$$C/O = \frac{\text{coke nake}}{\Delta \text{ coke}}$$

- 5 The present invention provides a process wherein lower boiling products are obtained from a hydrocarbon feed by contacting the feed with fluidized solid contact material at elevated temperature in a riser, spent contact material is separated and stripped of volatile hydrocarbons in a stripping zone, stripped contact material is
- 10 regenerated with oxygen-containing gas in a regeneration zone, and hot freshly regenerated fluidized solid contact material is returned to the riser, the process further including at least one, and preferably at least two, of the following :
- 15 (i) suspending hot regenerated contact material in a carbonizable lift gas at the lower portion of the riser and injecting hydrocarbon feed higher in the riser;
- 20 (ii) charging hydrocarbon feed into the base of the riser, steam cooling a portion of the hot freshly regenerated contact material, and contacting the hydrocarbon feed with hot regenerated contact material upsteam of where said hydrocarbon feed is
- contacted with the steam cooled contact material;
- (iii) cooling a portion of hot freshly regenerated contact material before returning it to the riser; and
- 25 (iv) adding a portion of freshly heated regenerated contact material to said stripping zone.

The process of the invention provides means to operate a heat balanced FCC unit or a selective vaporization unit at an increased C/O ratio. These means may be used alone or in combinations.

- 5 In accordance with one aspect of the invention, the CCR of an FCC unit, heavy oil FCC unit or selective vaporization process is controlled (increased) by directly cooling hot circulating fluid solid contact material. Preferably, hydrocarbon charge is fed directly to the base
10 of a riser along with lift gas and/or steam, if needed, and contacted with hot regenerated fluid solid contact material upstream of the point at which the gasiform mixture is contacted with cooled fluid solid contact material for increased C/O in a heat balanced operation without cooling
15 of the regenerator/burner system. This method of operation reduces thermal reactions and increases desired catalytic reactions in FCC units. This method of operation is also beneficial when feeds containing asphaltenes, basic nitrogen and metals, are being processed in FCC or
20 selective vaporization units. By the method, feed is preheated with a minimum of hot regenerated material to remove the aforementioned impurities and the cooled solid contact material injected immediately downstream of the hot solid contact material injection maintains a high
25 selectivity because active sites on the contact material are not covered with asphaltene (coke) deposits nor are they neutralized by basic nitrogen. Also, freshly deposited metals which are especially detrimental to activity have previously been removed during contact with hot regenerated
30 contact material.

1 In another aspect, the invention comprises a
system for controlling the operation of an FCC unit, heavy
oil FCC unit or a selective vaporization process to
increase C/O and superheat reactor/contacter vapors by
5 combining hot regenerated material with spent contact
material directly into the spent contact material stripper,
or in case of the apparatus, described hereinafter, by
combining hot regenerated material with reactor/contacter
products between the preseparator outlet and the high
10 efficiency cyclone inlet and returning the hot regenerated
material to the stripper through the high efficiency
cyclone dipleg. This will lower the Δ coke on the
circulating material which, in turn, will increase C/O
ratio by lowering the regenerator temperature. This
15 results in lower carbon on spent material by vaporizing
more of the hydrocarbon from the spent material in the
stripper. A secondary beneficial effect is that the
hydrocarbon vapors in the stripper and the high efficiency
cyclone inlet are heated to a higher temperature than they
20 would normally be heated. Since these vapors may be at
their dew point, any cooling will normally result in
condensation of the heavy ends, causing undesirable coke
formation in the vessel, cyclones and vapor lines. By
reheating these vapors, coke formation resulting from
25 condensation reactions is reduced.

In still another aspect of the invention C/O is
increased in an FCC unit or heavy oil FCC unit operating
with a zeolitic cracking catalyst by lifting regenerated
catalyst with a lift gas that is capable of being cracked
30 in a riser before regenerated catalyst contacts feed which
is injected downstream in the riser to control contact
time. The lift gas that is used forms coke on the acid
sites of catalyst before feed addition. As a result, the
acid sites of the catalyst are deactivated by carbon formed

1 when hot catalyst contacts lift gas. Preferred lift gases
are either wet gas from the main column overhead receiver
or any gas after recovery of C3's and C4's in a gas
concentration unit. The zeolitic sites are not deactivated
5 and are available to crack a gas oil feed. Coke is reduced
and yield structure is improved. This results in increased
octane and olefin production, as well as higher C/O.

In an especially preferred embodiment, the
process of this invention also employs controls on
10 both the reactor (contactor) and the regenerator
(combustor) so that there is complete control of the
circulating solid material. In the first place, when the
circulating material is in contact with combustion products
(regenerator/combustor) or hydrocarbon vapors
15 (reactor/contacter) it is in a dilute phase. After
separation of the circulating material from the vapors or
combustion products, the material is returned to a
different vessel. The vast majority of other technology
employed has a dense bed in contact with products of
20 combustion or hydrocarbon vapors and utilizes cyclones to
return the circulating material to the same vessel from
which it came. This vessel always contains a dense bed of
circulating material. In the novel process of this
invention, the dilute phase system is connected directly to
25 a preseparator and then to high efficiency cyclones such as
multicyclones so that the circulating material is always
discharged into another vessel separate from the vapors.
This circulating material forms a dense bed in the
secondary vessel, the secondary vessel being neither the
30 reactor/contacter or the regenerator. The process is
characterized by being a completely balanced system in that
the separation efficiency is the same for both the
reactor/contacter and the regenerator/combustor so that the
two systems will retain the same particle size range. In
35 the process, all the contact material entering the

1 reactor/contactor and regenerator/combustor is maintained
in a dilute phase and passes through the preseparator, i.
e., there is no dense fluid bed in either the
reactor/contactor or regenerator/combustor. The novel
5 process provides flexibility with respect to hydrocarbon
feedstock composition, product distribution throughput and
capacity. The process also results in more fines
retention which will result in lower average particle size,
more surface area, higher activity maintenance, and better.
10 fluidization properties.

Further features and advantages of the process of
this invention would be apparent from a detailed
description of the preferred embodiment of the process as
applied to the catalytic cracking of hydrocarbons.

15 The sole accompanying figure is a diagrammatic
representation of a preferred form of apparatus of the
present invention which is suitable for carrying out the
process of this invention.

With reference to the figure illustrating a
20 preferred embodiment of the catalytic cracking system of
this invention, the system comprises a fast fluid type
system that essentially operates in the dilute phase with
all of the material transported from the bottom of the
regenerator (B) to the top. The difference in this system
25 when compared with conventional systems is (1) complete
control of all catalyst flow into the system and (2) all
the catalyst that flows into the system is transported by
the air and eventually by the products of combustion
through the regenerator to the cyclones (E) and (F). In
30 this way, control is had not only of the total flow rate of
air and combustion products but also the catalyst loading
to the cyclone systems (E) and (F). The control of the
catalyst allows control of pounds per cubic foot of
catalyst entering the cyclones (E) and (F) and therefore
35 gives control of the loading to the cyclone so that the

1 system is not overloaded. Thus, air from a blower enters
through line (A) and through the bottom of regenerator (B)
containing catalyst to be regenerated and the flow rate is
controlled to maintain a dilute phase. The catalyst and
5 vapors are rapidly separated in preseparator (E) at a
efficiency greater than 80% and the catalyst material
passes through the bottom of preseparator (E) through line
(R) into the regenerator surge hopper (C). Vapor materials
exit (E) and pass into multicyclones (F) wherein flue gas
10 is removed through line (G) and catalyst again passes
through line (R) into regenerator surge hopper (C). The
regenerator system is obviously built to burn off all the
carbon from the spent catalyst. Therefore, this system is
designed for certain temperature, pressure, and flow rate
15 so that one can obtain the carbon burning in the desired
time. The products of combustion from the regenerator that
exit through flue gas line (G) are mainly nitrogen, CO₂,
with a small amount of CO, i.e., less than 500 parts per
million normally, SO_x which is dependent on the amount of
20 sulfur in the feed, water vapor contained in the combustion
air and water vapor produced by combustion reaction. There
is a small amount of NO_x produced that is strictly based on
the temperature of operation of the regenerator (B). Most
of the nitrogen in the coke that is related to the nitrogen
25 in the feed is liberated either as ammonia or as nitrogen.
This system can also be operated to leave carbon on the
regenerated catalyst by limiting the air to the
regenerator. This may result in higher CO levels in the
flue gas. The amount of CO will depend on the regenerator
30 temperature, carbon level on regenerated material and
oxygen supplied for combustion.

Since the spent catalyst is relatively cool,
i.e., between 800°F. and 1100°F., the burning of carbon
from this material is difficult at the time frame usually
35 employed for the dilute phase burning in the regenerator,

1 i.e., 3 to 15 seconds. For this reason, a hot
recirculation valve (O) is provided that circulates back
hot material from the regenerator surge hopper (C) back to
the base of regenerator (B). The purpose of this line is
5 to control the temperature in regenerator (B) so that the
carbon or coke can be burned off the spent catalyst in the
time allowed in the dilute phase transport riser
regenerator. Typically, the ratio of the circulation rate
through valve (O) and the circulation rate through spent
10 slide valve (JJ) is at least 1:1 and in many cases will be
2:1 or greater so that the temperature in this system can
be raised to that in which burning can take place
completely to CO₂ and all the carbon burned off the
catalyst. The figure also includes a preferred, though not
15 an essential, embodiment of the novel process of the
invention, namely catalyst cooler (N) whose flow rate is
controlled by slide valve (P). This valve is used when the
regenerator temperature reaches its maximum metallurgical
limits in order to protect the equipment against excessive
20 damages or to obtain a maximum temperature based on
catalyst activity maintenance or C/O consideration. In
other words, if it is desired to operate the regenerator at
a temperature of about 1600°F., when the temperature in the
system approaches that temperature the flow would be
25 started through slide valve (P) which would move more
material from regenerator surge hopper (C) through the
catalyst cooler (N) and through the valve (P). As is
obvious, this works opposite of valve (O) in that it would
be removing heat by generating steam in the catalyst cooler
30 (N) and therefore lower the overall system temperature in
the regenerator.

As can well be appreciated, both control valves
(P) and (O) can be used. If one wishes to produce steam in
order to supply energy to a refinery, valve (P) can be used
35 continually in the open mode to generate a constant amount

1 of steam. There may also be reasons to operate at cooler
temperatures in the regenerator because catalyst activity
maintenance, catalyst to oil relationships in the reactor
or contactor (K), or due to the desired reaction kinetics.
5 As indicated earlier, at the top of transport riser
regenerator (B) all of the catalyst and air used for
combustion enters preseparator (E). Preseparator (E) is
designed so that the minimum efficiency must be greater
than 80% and preferably greater than 90% removal of the
10 solids from the gas. The solids removed are discharged
from separator (E) through regenerated solids line (R) to
the regenerator surge hopper (C). This return of the
material from the preseparator (E) to the surge hopper (C)
is done at a level lower than the bed level in surge hopper
15 (C) so that the pipe is submerged in an actual level to
prevent back flow of gas up the pipe into preseparator (E)
causing preseparator (E) to malfunction. The flue gas that
exits preseparator (E) has only 20% or less of the catalyst
with which it came into preseparator (E) and it flows out
20 to the high efficiency cyclone system (F) for final clean
up. The total clean up in this system is greater than
99.0%, and preferably greater than 99.99%, and the
essentially catalyst-free gas exits high efficiency cyclone
(F) through line (G) to flue gas treating and/or the
25 atmosphere. The catalyst that is separated in high
efficiency cyclone (F) is returned to the regenerator surge
hopper (C) again below the normal level of catalyst so that
this dipleg is sealed.

The regenerator surge hopper (C) is fluidized by
30 controlling a small amount of air from the air blower into
the system through an air distributor so that the material
is maintained at at least the velocity of about one-half
foot a second but no greater than 3.3 ft. a second in the
vessel. The gas carrying some catalyst exits through
35 regenerator hopper (C) through line (L) which is connected

1 back to the regenerator (B). Line (L), the surge hopper
vent, is also an equalizing line and is a very key feature
of this process. This line assures that the pressure at
the inlet to preseparator (E) and the pressure on surge
5 hopper (C) are equal so at no time can surge hopper (C) be
at a much higher pressure than preseparator (E). If the
surge hopper (C) were at a higher pressure than
preseparator (E) then it would be possible that the
material that was separated in preseparator (E) could not
10 flow down pipe (R) into the surge hopper and therefore
cause preseparator (E) not to function. The same pressure
differential or equalization is necessary from high
efficiency cyclone (F) to surge hopper (C). Surge hopper
(C) can be at a slightly higher pressure than either (E) or
15 (F) as long as the level in the return pipes (R) is not
high enough to cause preseparator (E) and high efficiency
cyclone (F) to malfunction. The vent line (L) is shown
connected to the dilute phase regenerator (B). However, the
vent line (L) from the regenerator inventory surge hopper
20 (C) could also be placed between (E) and (F) and that
choice depends on the velocity component in regenerator
surge hopper (C). In any event, the purpose of vent line
(L) is to insure that the pressure at inlet to preseparator
(E) and the pressure on surge hopper (C) are equal.

25 Similar considerations apply to reactor/contacter
(K) through which is introduced lift steam through line (J)
and hot regenerated catalyst through slide valve (Q) and
feed through line (HH) and the products and catalyst again
empty into preseparator (E) and then into (F) in the same
30 manner as has been previously described with respect to
regenerator (B). It is noted that it is also essential
that there be an equalizer line (M) from the catalyst
stripper (D) back to the cyclones (E) and (F) in the
identical same manner as there is between regenerator
35 inventory surge hopper (C) and regenerator (B). Therefore

1 line (M) in the drawing performs the same function as line
(L).

Typically, the operating parameters for
regenerator (B) will be between 1100°F. and 2000°F. and for
5 FCC operations less than 1400°F. The velocity must be
greater than 3-1/2 ft. per second in order to assure a
dilute phase operation and less than 100 ft. per second and
usually will be maintained in the range of 5-15 ft. per
second. The pressure on the regenerator will typically be
10 between 5 and 50 psig, preferably between 10 and 30 psig
and gas time will typically be between 3 and 15 seconds.

The reactor system design is very similar to that
of the regenerator in the figure in that it also consists
of a two separator system (E) and (F) and a stripper (D)
15 which also functions as a surge hopper as well as a
riser/contactor (K). The regenerated catalyst is taken
from the regenerator surge hopper (C) through valve (Q)
into the riser contactor (K). It can be contacted with a
diluent such as lift gas, steam, hydrocarbon recycle, or
20 water; or be fed through line (J). There are also
provisions (HH) for an optional feed point to control the
time in the contactor or reactor. The regenerated catalyst
plus any diluent plus any recycle plus feed is contacted in
contactor/reactor (K) for a period of time necessary to
25 obtain the desired yield as either an FCC, heavy oil type
FCC, fluid coker or as in an ART (selective vaporization)
process, such as that disclosed in U. S. 4,263,128. Again,
all of the catalyst as well as all of the vapors produced
in reactor/contactor (K) enter into preseparator (E). As in
30 like manner with regenerator (B), the catalyst and vapors
to preseparator (E) are controlled so that the system is
not overloaded. The efficiencies of separators (E) and (F)
are as discussed on the regenerator system and again the
catalyst separated from the vapors is returned through
35 lines (R) below the dense bed level to spent catalyst

1 stripper (D). The spent catalyst stripper (D) is fluidized
with steam to the stripper. The vent line (M) from
stripper (D) enters between the separators (E) and (F).
Since the amount of entrained material will be quite low,
5 it could just as easily enter into the inlet to (E) as in
the regenerator.

The spent catalyst then leaves the catalyst
stripper on level control through valve (JJ). The vapors
now essentially free of catalyst leave the system through
10 line (H) to fractionation and to separation. In the case
of an ART unit they could be quenched at this point. In
the case of a fluid coker, FCC or heavy oil FCC, the vapors
would go into the fractionation system and may or may not
be quenched.

15 From the above description it should be noted
that one of the main differences between the system set
forth in the accompanying figure and the system of the
prior art is that none of the vapors from regeneration
system (B) or the contactor (K) are in equilibrium or
20 contact with the catalyst dense bed that would be contained
in vessels (C) or (D). Most systems up to this point have
a feature where the vapors are separated in the vessel and
are in constant contact with the dense bed.

The contactor (K) conditions are basically
25 between 10 and 100 ft. per second and preferably running at
an outlet velocity of about 70 ft. per second. The time
depends on whether there is an ART unit; as an ART unit,
the time would be preferably less than a second, and
normally less than 3 seconds, or as an FCC, which normally
30 operates between 1 second and 5 seconds vapor time. The
temperature in the contactor would range between 800 and
1000°F.

The preseparator (E) is not narrowly critical and
all that is required is that there be a very rapid
35 disengagement of circulating solids and vapors. Materials

1 of this type are disclosed in U. S. 4,285,706; U. S.
4,348,215; and U. S. 4,398,932, the entire disclosures of
which are herein incorporated by reference. The high
efficiency cyclone (F) is a conventional type cyclone as to
5 be understood that it can be one or a plurality of
cyclones. Preferred separation (F) is of the multicyclone
type, described in U. S. 4,285,706; the disclosure of which
is incorporated herein.

Figure 1 also depicts systems for controlling an
10 FCCU, heavy oil FCCU or ART Process to increase C/O by:
directly cooling the circulating catalyst by using a
cooler (V) and slide valve (W); lowering the carbon on the
circulating material (catalyst or catalytically inert
contact material in the case of an ART unit) which in turn
15 will increase the C/O ratio by lowering the regenerator
temperature by reheating the material in the
reactor/contacter stripper. This is accomplished by
combining hot regenerated material directly to the stripper
through slide valve (S) and heating riser (U). Lift media
20 (T) can be either gas or steam. Not only does this result
in lower carbon on spent material by vaporizing more of the
hydrocarbon from the material in the stripper, it has a
secondary beneficial effect of superheating the CX/RX
vapors which will reduce coke formation through
25 condensation reactions. Since the CX/RX vapors in heavy
oil FCC or an ART unit may be at their dew point, any
cooling results in condensation of the heavy ends which
results in coke formation in the vessel, cyclones and vapor
line. This superheating of the vapors will eliminate this
30 problem. As shown in the figure an alternative line (U¹)
can also be used to superheat the CX/RX vapors and increase
the stripper temperature by injecting hot regenerated
material and lift media (T) into the outlet of preseparator
(E).

1 Another feature shown in the figure is using lift
gas (J) to lift the regenerated catalyst before contact
with the feed (HH) which is injected higher in the riser to
control time in the contactor on reactor riser. This lift
5 gas is used to form coke on the acid sites before feed
addition to reduce coke formation and improve yield
structure. In conventional catalyst systems the acid site
activity greatly increases as the catalyst is regenerated
to carbon levels less than 0.3. This increased acid site ..
10 activity increases coke formation of the hydrocarbon
feedstock and reduces selectivity. Therefore, by
contacting the catalyst with gas before feed injection the
acid sites are deactivated by carbon formation from the
gas and the zeolitic sites, which are selective, are
15 available for cracking the gas oil. This results in
increased octane, olefins, and higher C/O.

 Still another feature shown in the figure is the
option of putting the hydrocarbon feed directly into the
base of the riser at (J) along with lift gas or steam and
20 contacting it with hot regenerated catalyst before
contacting it with cooled catalyst for increased C/O over a
heat balanced operation without cooling. This method of
operation is beneficial when feeds containing asphaltenes,
basic nitrogen and metals are being processed in an FCC
25 system. Using this method allows the operator to first
preheat the feed with a minimum of hot regenerated material
to remove the majority of the asphaltenes, basic nitrogen
and metals so that the cooled catalyst injected just
downstream of the first hot catalyst maintains a high
30 selectivity because the active sites are not covered by
asphaltene (coke) deposits, neutralized by basic nitrogen,
or competed with by fresh metals activity.

1 On skilled in the art of fluidized solid
handling would quickly realize the potential of the
illustrated transport system to reduce capital costs
associated with the installation of this equipment as well
5 as its potential to increase (double) the capacity of
existing FCC, heavy oil FCC, fluid coker or ART process
units using the existing vessels without major
modifications since this system eliminates superficial
velocity constraints now in existence with dense bed type
10 regenerators. Also, this fluidized apparatus design
concept is applicable to other fluidizable solids systems
which have heretofore been limited by superficial bed
velocities.

It is understood that the particular fluidized
15 solid or solids obviously depends on the particular process
being carried out.

Typical solids for cracking include those which
have pore structures into which molecules of feed material
may enter for adsorption and/or for contact with active
20 catalytic sites within or adjacent to the pores. Various
types of catalysts are available within this
classification, including for example the layered
silicates, e.g. smectites. Although the most widely
available catalysts within this classification are the
25 well-known zeolite-containing catalysts, non-zeolite
catalysts are also contemplated.

The preferred zeolite-containing catalysts may
include any zeolite, whether natural, semi-synthetic or
synthetic, alone or in admixture with other materials which
30 do not significantly impair the suitability of the
catalyst, provided the resultant catalyst has the activity
and pore structure referred to above. For example, if the
virgin catalyst is a mixture, it may include the zeolite
component associated with or dispersed in a porous
35 refractory inorganic oxide carrier, in such case the

1 catalyst may for example contain about 1% to about 60% more
preferably about 15 to about 50%, and most typically about
20 to 45% by weight, based on the total weight of catalyst
5 (water free basis) of the zeolite, the balance of the
catalyst being the porous refractory inorganic oxide alone
or in combination with any of the known adjuvants for
promoting or suppressing various desired and undesired
reactions. For a general explanation of the genus of
zeolite, molecular sieve catalysts useful in the invention;
10 attention is drawn to the disclosures of the articles
entitled "Refinery Catalysts Are a Fluid Business" and
"Making Cat Crackers Work on Varied Diet", appearing
respectively in the July 26, 1978 and September 13, 1978
issues of Chemical Week magazine. The descriptions of the
15 aforementioned publications are incorporated herein by
reference.

For the most part, the zeolite components of the
zeolite- containing catalysts will be those which are known
to be useful in FCC cracking processes. In general, these
20 are crystalline aluminosilicates, typically made up of
tetra coordinated aluminum atoms associated through oxygen
atoms with adjacent silicon atoms in the crystal structure.
However, the term "zeolite" as used in this disclosure
contemplates not only aluminosilicates, but also substances
25 in which the aluminum has been partly or wholly replaced,
such as for instance by gallium, phosphorus, boron, iron,
and/or other metal atoms, and further includes substances
in which all or part of the silicon has been replaced; such
as for instance by germanium or phosphorus, titanium and
30 zirconium substitution may also be practiced.

Most zeolites are prepared or occur naturally in
the sodium form, so that sodium cations are associated with
the electronegative sites in the crystal structure. The
sodium cations tend to make zeolites inactive and much less
35 stable when exposed to hydrocarbon conversion conditions,

1 particularly high temperatures. Accordingly, the zeolite
may be ion exchanged, and where the zeolite is a component
of a catalyst composition, such ion exchanging may occur
before or after incorporation of the zeolite as a component
5 of the composition. Suitable cations for replacement of
sodium in the zeolite crystal structure include ammonium
(decomposable to hydrogen), hydrogen, rare earth metals,
alkaline earth metals, etc. Various suitable ion exchange
procedures and cations which may be exchanged into the
10 zeolite crystal structure are well known to those skilled
in the art.

Examples of the naturally occurring crystalline
aluminosilicate zeolites which may be used as or included
in the catalyst for the present invention are faujasite,
15 nordenite, clinoptilolite, chabazite, analcime, erionite, as
well as levynite, dachiardite, paulingite, noselite,
ferrierite, heulandite, scolecite, stibite, harmotome,
phillipsite, brewsterite, flarite, datiolite, gmelinite,
caumite, leucite, lazurite, scapolite, nesolite, ptoleite,
20 nepheline, natrolite, offretite and sodalite.

Examples of the synthetic crystalline
aluminosilicate zeolites which are useful as or in the
catalyst for carrying out the present invention are Zeolite
X, U. S. 2,882,244; Zeolite Y, U. S. 3,130,007; and Zeolite
25 A, U. S. Patent No. 2,882,243; as well as Zeolite B, U. S.
Patent No. 3,008,803; Zeolite D, Canada Patent No. 661,981;
Zeolite E, Canada Patent No. 614,495; Zeolite F, U. S.
Patent No. 2,996,358; Zeolite H, U. S. Patent No.
3,010,789; Zeolite J, U. S. Patent No. 3,011,869; Zeolite
30 L, Belgian Patent No. 575,177; Zeolite M, U. S. Patent No.
2,995,423; Zeolite O, U. S. Patent No. 3,140,252; Zeolite
Q, U. S. Patent No. 2,991,151; Zeolite S, U. S. Patent No.
3,054,657; Zeolite T, U. S. Patent No. 2,950,952; Zeolite
W, U. S. Patent No. 3,012,853; Zeolite Z, Canada Patent No.
35 614,495; and Zeolite Omega, Canada Patent No. 817,915.

- 1 Also ZK-4HJ, alpha beta and ZSM-type zeolites are useful.
Moreover, the zeolites described in U. S. Patent Nos.
3,140,249; 3,140,253; 3,944,482; and 4,137,151 are also
useful, the disclosures of said patents being incorporated
5 herein by reference.

The crystalline aluminosilicate zeolites having a
faujasite-type crystal structure are particularly preferred
for use in the present invention. This includes
particularly natural faujasite and Zeolite X and Zeolite Y.

- 10 Typical solids for the ART process are those set
forth in U. S. Patent No. 4,263,128.

- It is to be emphasized that whilst the invention is described and
14 illustrated above mainly in terms of a cracking process in which the
contact material is a cracking catalyst and the riser is a reactor,
it applies similarly to a selective vaporisation process in which
the contact material is substantially inert catalytically and the
riser is termed a contactor. Whether the contact material is
catalytic or inert, it does tend to become spent in the riser and
20 is regenerated for recycling by combustion of deposits in the
regeneration zone; terms such as "regeneration zone", "regenerator"
etc. are appropriate for both FCC and selective vaporisation
systems, though herein terms such as "combustion zone", "combustor"
etc. are sometimes used instead in connection with the selective
25 vaporisation system.

GLOSSARY FOR FIGURE

- A. Air Blower Discharge - Air to Regenerator System
- B. Regenerator/Burner/Kiln
- C. Regenerator/Surge Hopper
- D. Spent Catalyst/ARTCAT Stripper
- E. Preseparator
- F. High Efficiency Cyclone
- G. Flue Gas to Stack/Treating
- H. Oil Vapors to Separation
- I. Steam to Stripper
- J. Lift Steam/Gas (Wet or Dry)/H₂O to Riser Contactor or Feed
- K. Riser Contactor
- L. Surge Hopper Vent
- M. Stripper Vent
- N. Catalyst Cooler/ARTCAT Cooler
- O. Hot Recirculating Material Slide Valve
- P. Cold Recirculating Material Slide Valve
- Q. Regenerated Material Slide Valve
- R. Separated Material from Separators
- JJ. Spent Slide Valve
- HH. Feed - Optional Feed Point to Control Time in Contactor
- S. Hot Material to Cx/Rx Heating Riser Slide Valve
- T. Lift Gas/Steam to Cx/Rx Heating Riser
- V. C/O Control Cooler
- W. Cooled Catalyst to Riser Contactor
- U. Cx/Rx Stripper Reheat Line
- U¹. Cx/Rx Vapor Superheat Line

C L A I M S

1. A process wherein lower boiling products are obtained from a hydrocarbon feed by contacting the feed with fluidized solid contact material at elevated temperature in a riser, spent contact material is separated and stripped of volatile hydrocarbons in a stripping zone, stripped contact material is regenerated with oxygen-containing gas in a regeneration zone, and hot freshly regenerated fluidized solid contact material is returned to the riser, the process further including two or more of the following :

- (i) suspending hot regenerated contact material in a carbonizable lift gas at the lower portion of the riser and injecting hydrocarbon feed higher in the riser;
- (ii) charging hydrocarbon feed into the base of the riser, steam cooling a portion of the hot freshly regenerated contact material, and contacting the hydrocarbon feed with hot regenerated contact material upstream of where said hydrocarbon feed is contacted with the stream cooled contact material;
- (iii) cooling a portion of hot freshly regenerated contact material before returning it to the riser; and
- (iv) adding a portion of freshly heated regenerated contact material to said stripping zone.

2. A process according to claim 1 wherein the contact material comprises zeolite and contains acid sites.

3. A process according to claim 1 or 2 wherein the contact material is a cracking catalyst comprising zeolite and contains acid sites, and the feed contains asphaltenes, basic nitrogen and metals, and method (i) is one of the methods used.

4. A process according to any preceding claim wherein the contacting is carried out in a dilute phase and all of the solid material from the riser is subjected to cyclone pre-separation with rapid disengagement of solids and gases, the separated solids being returned to a dense bed in a vessel other than the riser.

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5. A process according to any preceding claim wherein the regeneration is carried out in a dilute phase and all of the solid material from the regeneration zone is subjected to cyclone pre-separation with rapid disengagement of solids and gases, the separated solids being returned to a dense bed in a vessel other than the regeneration zone.

6. A process according to claim 1 and substantially as hereinbefore described with reference to the accompanying drawings.

